Energy Conversion and Management 51 (2010) 1270–1276

Contents lists available at ScienceDirect

Energy Conversion and Management

journal homepage: www.elsevier.com/locate/enconman

Energy consumption for sugar manufacturing. Part I: Evaporation versus reverse osmosis

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ARTICLE INFO

Article history:
Received 3 September 2008
Received in revised form 12 June 2009
Accepted 9 January 2010
Available online 7 February 2010

Keywords:
Reverse osmosis
Evaporation
Sugar thin juice concentration
Energy

ABSTRACT

Removing water from various feeds is usually carried out using evaporation process especially in food industry. Due to the high latent heat of water, this unit operation results in consumption of unacceptable amount of energy. Finding low energy consuming processes which could be replaced with this process is still a challenge.

The processes with no phase inversion may be considered for concentration purposes with reasonable energy consumption in comparison with the other various separation procedures. Reverse osmosis and most of the other membrane technologies are separation techniques without any change in the phase and therefore consume low amount of energy.

Concentrating the sugar thin juice in the classical sugar manufacturing procedure is carried out using conventional evaporation. Reverse osmosis membranes may be used as a pre-concentration step to partially separate water from the sugar thin juice in combination with this part of the plant. Final concentration and thick juice preparation for crystallization may be carried out in the evaporation unit.

In this study, membranes were employed for sugar thin juice concentration using a two-stage reverse osmosis process in two different arrangements. The energy consumption was calculated and compared for conventional evaporation versus reverse osmosis combined with evaporation. The results indicate that the employment of revers osmosis membranes for concentrating the sugar thin juice leads to sensibly lower energy requirements. Furthermore, there is no thermal loss of sugar in the membrane process.

1. Introduction

Evaporation is one of the most energy intensive unit operations in sugar factories. After purification of sugar thin juice in filters, the feed enters into the multi-effect evaporator with the brix degree (percent of dry substance in the solution) of around 15. The brix degree of the concentrated sugar solution is more than 60. Due to the high latent heat, evaporation of water consumes huge amount of thermal energy and fuel which results in higher operating costs and environmental problems. Furthermore, heating sugar juice could lower the product quality by changing the color and flavor.

Always finding low energy consuming alternative processes have been a challenge for scientists. However most of the researches have focused on the energy optimization of current processes or looking for other fuel choices with lower costs and pollutions. Kilicaslan et al. investigated the possibility of using bagasse, obtained in the processing of sugar-cane, as a fuel for boilers in an effort to decrease consumption of fossil fuels [1].

The use of cogeneration in the sugar-cane industry is a common practice all over the world, although until recently the electrical energy produced was consumed within the plant. In some countries such as Brazil and Cuba, two of the largest producers of sugar-cane, the possibility of supplying electricity to the grid by means of cogeneration in the sugar-cane industry has been the subject of discussion in the last years. A thermo-economic analysis of a Cuban sugar-cane mill was investigated by Campo et al. for this purpose [2]. Also a detailed analysis of process steam demand reduction and electricity generation in sugar and ethanol production from sugar-cane was performed by Ensinas et al. They thoroughly investigated different configurations to save more energy and increase surplus electricity generation, and showed that this increment could be higher than 70% in some cases [3].

The traditional beet sugar manufacturing technology has a considerable detrimental impact on the environment. These problems can only be partly solved through a rational use of the by-products and improving the plants use of traditional technology. To make major improvements and to produce an optimal solution for the problem, innovative technologies should be exploited [4]. One of the best candidate processes to replace traditional evaporation or optimize its energy consumption is membrane process with no

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doi:10.1016/j.enconman.2010.01.002

Nomenclature

Latin notation	
Bx	brix degree
C	consumption of thermal primary energy (MJ or Mcal)
CS	substitution coefficient (MJ/kW h)
E	consumption of electrical energy (kW h)
m_S	mass flow rate of steam (kg/s)
P_{max}	maximum pressure of pump (bar)
q	thermal power (kW)
Q	flow rate (m^3/s)
R	rejection (%)
SG	specific gravity
T_{max}	maximum safe temperature of pump (°C)
W	power (W)
Greek symbols	
η	efficiency of the pump
λ	latent heat of water evaporation (kJ/kg)
ΔP	pressure difference (Pa)
Subscript	
F	feed
P	permeate
R	retantate

phase inversion included. Membrane processes are very energy effective and could replace current energy intensive processes like evaporation and distillation solely or in hybrid configuration. Indeed the energy crisis and environmental aspects make it critical to take a wider look to basic changes toward new processes instead of minor changes in the fuel type or details of the current processes.

Membranes have the ability of separating water from sugar. Some of the membrane processes such as electrodialysis [5], microfiltration [6–11], ultrafiltration [7–33] and nanofiltration [7,20,34,35] could be utilized to refine sugar juice. Also membrane distillation [36], osmotic distillation [37–43], nanofiltration [14,30,33,44,45] and reverse osmosis could be applied to sugar thin juice concentration.

Reverse osmosis possible application in sugar industry is a known subject for researchers [25]. Bichsel et al. [46] carried out a research to concentrate sugar thin juice from 13% to 30% of sugar content using PA300 and FT 30 reverse osmosis membranes. They showed the advantages of this method over evaporation including lower cost. There are some literatures about using combined UF-RO systems for thin juice concentration [47]. Kiss et al. [48] compared a MF-RO-NF membrane process with traditional evaporation in must production. The goal of direct production of white cane sugar by means of clarification and decolorization membranes is followed by Saska et al. [44] in his article reporting about concentration and decolorization of dilute products from cane molasses desugarization with RO and NF membranes. Hogan et al. [39] reported that a hybrid process involving pre-concentration of the feed by RO followed by further concentrating the RO retentate by osmotic distillation, should yield a high solids product concentrate of an excellent quality, but at significant reduction in processing cost [39]. For instance, this reduces the evaporator capacity requirements by one-half. However there are some works showing the possibility of using membranes in sugar industry, actually there are few studies discussing energy saving aspects quantitatively.

In a previous work [49] we studied the application of various commercial reverse osmosis membranes in several operating pressures for sugar syrup concentration. One commercial nanofiltartion membrane (NF45) and five commercial reverse osmosis membranes (DS, DSII, PVD, FT 30 and BW30) were evaluated in different operating pressures to concentrate thin sugar juice from around 15% sugar to a higher concentration. The results showed that nanofiltration NF45 membrane has no effect on sugar syrup concentration. The rejections of sugar using DSII and PVD reverse osmosis membranes vary between 23% and 33% for different operating conditions. DS membrane rejected around 10% of the sugar molecules in the best conditions. FT 30 membrane initially showed superior performance (55%). However, the rejection was decreased during time to 7%. For BW30 membrane, the rejection of sugar was better (60%) compared to the other membranes used in that study. For two-stage processes (i.e. the permeate of the first stage used as a feed for the second stage) the highest rejection (88%) was obtained. Therefore we concluded that BW30 reverse osmosis membrane at 22 bars is the most promising membrane, among those tested, for concentrating sugar thin juice. Note that there is always a limitation in the feed concentration, since the osmotic pressure rises progressively with respect to concentration which means higher energy consumption in RO process [50]. In the present study we investigated and quantified the energy consumption for sugar syrup concentration using reverse osmosis membrane as a pre-concentration versus evaporation.

2. Materials and methods

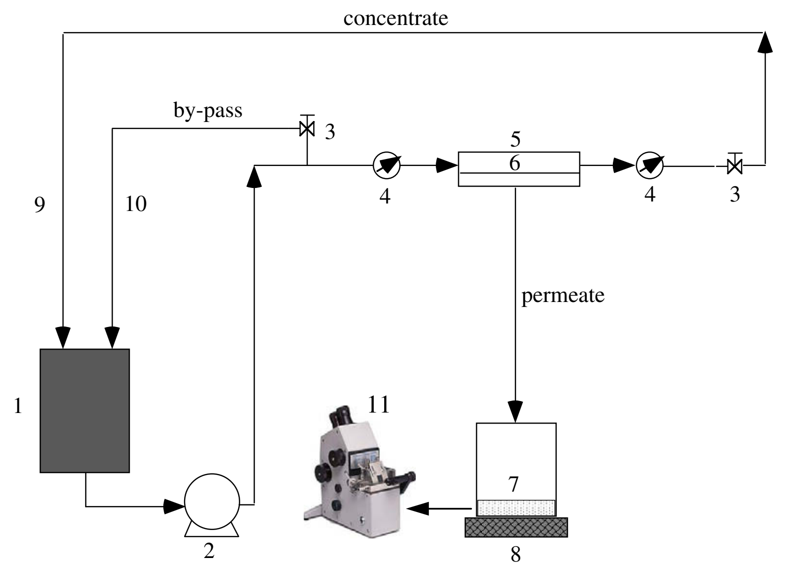
2.1. Experimental setup

A bench-scale cross-flow batch concentration process (Fig. 1) was used for all experiments. Permeate was taken out of the loop and concentrate was recycled to the feed tank. The high-pressure pump was a reciprocating type with P_{max} = 110 bar. Its maximum safe temperature was T_{max} = 50 °C. Connector hoses were of high-pressure type. There was a bypass valve and a valve on concentrated stream to control the applied pressure. These valves are used to adjust the flow rate and operating pressure over the membrane surface.

The membrane cell consisted of two cubic parts which made of a specific strong alloy. The RO membrane with the area of 0.0023 m^2 was sandwiched between the two parts and settled on a resistant compact foam layer to protect it against deformation and displacement. There were two oil pressure gauges (0–60 bar) before and after the cell which enable us to measure the pressure of the concentrated phase and the pressure drop.

2.2. Membrane

The BW30 reverse osmosis membrane manufactured by DOW with the general trademark of FILMTEC was used for all experiments. BW stands for brackish water and BW membranes are usually used for desalination purposes. This membrane is a commercial polyimide thin-film composite membrane with a smooth surface. The skin material makes this membrane hydrophobic. The BW30 membrane has a thickness of 150 μm. The maximum operating temperature and pressure of Filmtec modules are 45 °C and 41 bars, respectively [51]. Most of the scientists believe that RO membranes are non porous and cut-off is not applicable for RO membranes. Nevertheless due to the official website of the manufacturer, the MWCO of RO membranes is around 100 Da.



1	Feed tank (Plastic, 15 l)	7	Permeated water
2	Feed pump (Bertolini KKL 3021)	8	Balance (KERN EWB 220-2M, 0-220±0.01g)
3	Valve	9	Concentrate stream
4	Pressure gauge (PAKKENS, EN 837-1, 0-60±1 bar)	10	By-pass stream
5	Crossflow membrane cell	11	Refractometer (Bellingham&Stanley Ltd., Abbe 60/DR, 0-95±0.5 Brix)
6	Membrane (BW30)		

Fig. 1. Experimental set-up.

2.3. Experimental procedure

Experiments were carried out with a specified amount of thin juice (about 10 l) which has permeated through the filters and is to be evaporated. The initial brix of this feed was around 15°. Due to the maximum safe temperature of the pump (50 °C), the thin juice was cooled down from 80 °C to 30 °C to protect the pump against overheating during the experiment. A piece of membrane with the appropriate area was cut out to fit the cell and before the experiment soaked in a 50% solution of water and ethanol for 10–15 min to make it hydrophilic. Sampling from permeated stream and measurements were started immediately after running the feed pump and adjusting the feed pressure. The permeated liquid was collected in small scaled glassware during specified time intervals and the volume and weight were measured (KERN EWB 220–2 M, ±0.01 g) to calculate the flux.

Brix degree is linearly dependent on the sugar concentration. The brix degree of the permeated water and the feed tank were measured with a refractometer (Bellingham & Stanley Ltd., Abbe 60/DR, ±0.5° brix). Note that due to the low permeate flow rate, the brix degree variations in the feed tank was negligible. The rejection of membrane was calculated using the juice brix degree as follows:

$$R = (1 - Bx_P/Bx_F) \times 100 \quad (1)$$

where R is rejection (%), Bx is brix degree and P and F are subscripts indicating permeate and feed streams, respectively. The experiment was also carried out with a feed of brix 20°. The permeate of the experiments (with a low sugar content) was passed through a second stage to achieve a sugar free permeate. Fluxes and rejections were calculated for all the experiments.

3. Results and discussion

Fluxes and rejections obtained in the experiments are represented in Fig. 2. Note that uncertainty values for flux is less than one (±0.2) which could not be represented on the graphs. However these values for rejections, calculated with respect to the accuracy of refractometer, depend on brix range (e.g. ±3 for brix 20) and vary on different graphs. We calculated the energy consumption based on the average of achieved flux. We considered minimum efficiency and maximum energy consumption for membrane system in whole calculations. The feed side pressure is over 22 bar to provide the transmembrane pressure. This pressure should be over the osmotic pressure (Fig. 3).

The input sugar juice flow rate in a typical Iran beet sugar factory (e.g. Bisotoun Sugar Factory – Kermanshah, Iran) is about 80 m^3/h. The brix of second stage concentrate stream is 15° which is recycled to the main feed and the main concentrate stream has a brix of 20°. Based on the obtained results, the permeate flux for the first stage was considered as 10 l/m^2 h with brix degree of 4 which is fed into the second stage and for the second stage as 20 l/m^2 h with brix degree of 1. Note that the permeate brix degree for a feed with brix degree = 2–5 is about 0.2–0.3 which can be considered as pure water. Fig. 4 represents a general schematic of the two-stage process.

3.1. Arrangement one: booster pumps and equal number of parallel modules

In this arrangement sugar thin juice is pumped into the first stage by a main pump with a pressure slightly more than experimental pressure. Due to the pressure drops this pressure considered to be 25 bar. Then this feed is distributed between the first set of parallel modules according to the allowed feed flow rate of each module. The concentrate of this set enters to the second set which has the same number of modules as the first one. Assuming the pressure drop of each module to be 1 bar, after five set of parallel modules a booster pump compensates this pressure drop and raises back the pressure from 20 bar to 25 bar. The other modules are also set in this way, and the last set of modules consists of the remained modules which are the same or less than number of

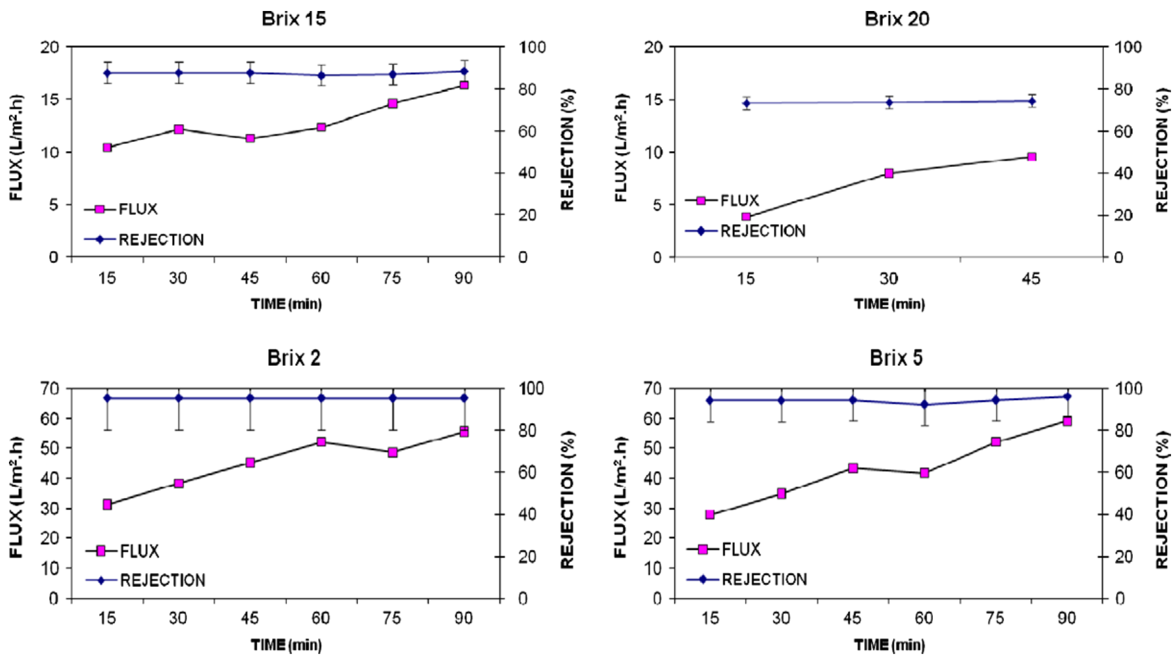


Fig. 2. Fluxes and rejections obtained in the experiments.

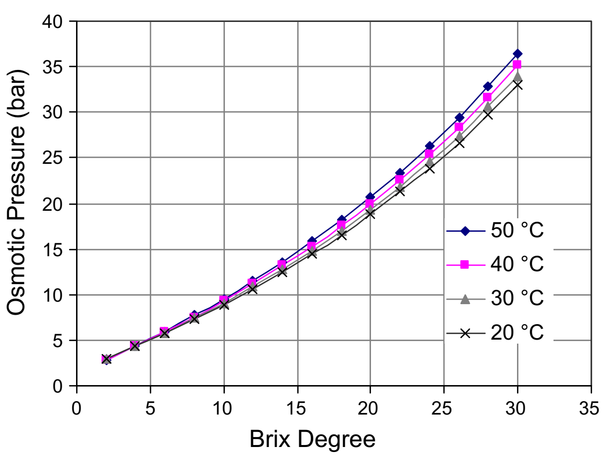


Fig. 3. Osmotic pressure of sugar juice versus brix degree at different temperatures [50].

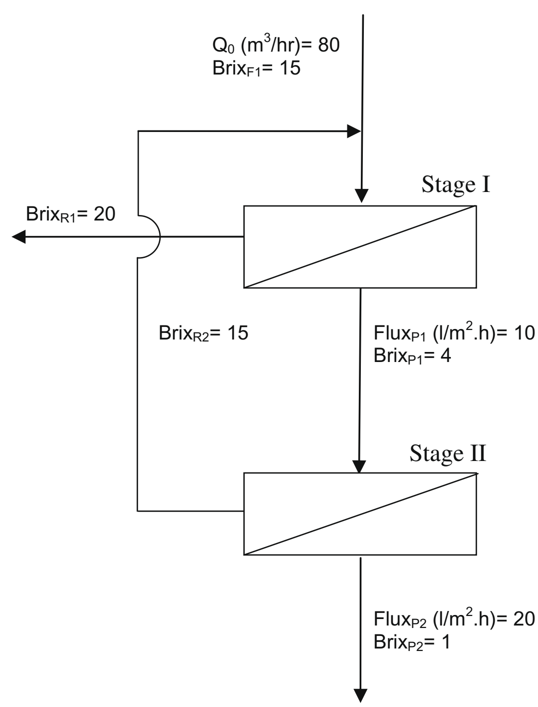


Fig. 4. General schematic of the two-stage membrane process.

other sets. The concentrate of this last set is the main concentrate stream of the RO system. The permeate stream of all modules of the first stage is fed to the second stage as a single stream.

The second stage have the same module arrangement as the first one, but the total number of modules and the number of modules in parallel sets could differ from the first stage. The last concentrated stream recycles to the main feed of the first stage and the permeate stream of all modules provides the main permeate stream of the RO system as removed water. An overview of this arrangement is represented in Fig. 5.

3.2. Arrangement two: single pumps stage and decreasing number of parallel modules

In this arrangement which is like a "Christmas tree", dilute feed is pumped into the first stage by the main and only pump of this stage with a pressure more than experimental pressure and even more than the first arrangement operating pressure. Because there is no booster pump to compensate pressure drops. This pressure considered to be 28 bar. Then this feed is distributed between the first set of parallel modules. The membrane area and the number of modules is this arrangement is equal to the other arrangement. So the number of these first modules is more than the first set in the first arrangement. The second set of parallel modules consists of fewer modules than the first set.

We considered three sets of parallel modules in each stage of this arrangement. As in the first arrangement, the final concentrate of the first stage is the concentrated product and the permeate of the first stage modules make the second stage feed.

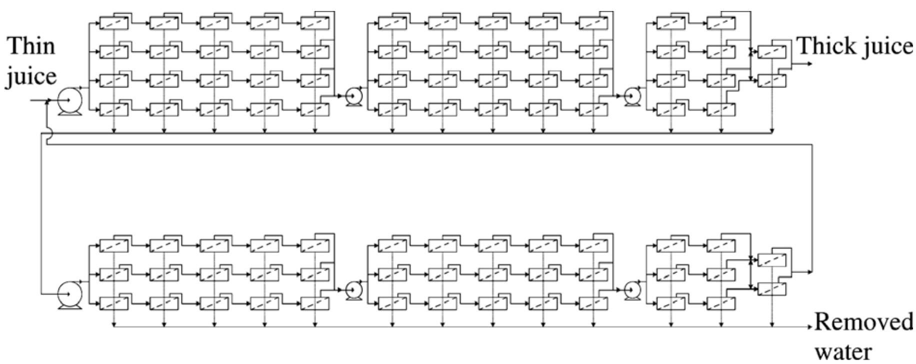


Fig. 5. A sample overview of the first arrangement.

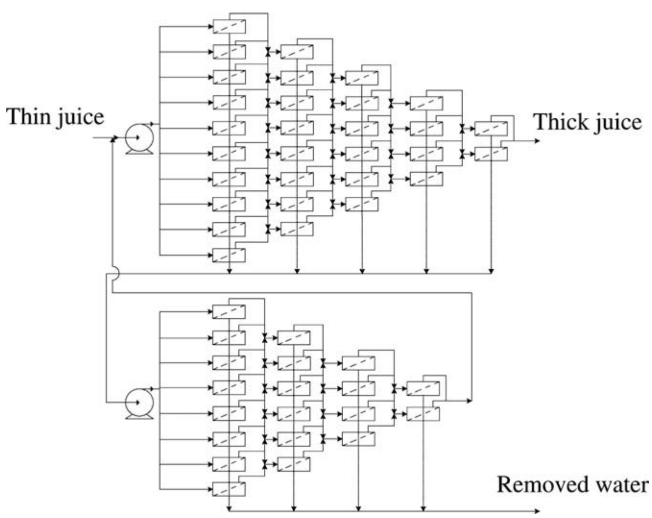


Fig. 6. A sample overview of the second arrangement.

The total number of modules and the number of modules in parallel sets in the second stage differ from the first stage, but the overall arrangement is the same. The concentrated stream of this stage is recycled to the first stage and the permeates of all modules exit as removed water. Fig. 6 shows an overview of the second arrangement.

3.3. Membrane area and energy calculations

A material balance for sugar and water based on obtaining output sugar juice with the brix of 20°, leads to the flow rate of the juice and the required membrane area. To convert volume and mass we need the relation between specific gravity and brix degree, which at 20 °C is approximately [52]:

$$SG = 0.00425Bx + 0.9988 \quad (2)$$

Our calculations showed that the needed membrane areas are 2801.38 m^2 and 1114.53 m^2 for stage one and two respectively. Using Filmtec™ BW30-400 standard spiral wound modules with the active area of 400 ft^2 (37 m^2), we need 76 and 31 modules for the first and second stage respectively [51]. The maximum feed flow of these modules is 19 m^3/h (5.3 l/s). We considered the feed flow rate of each module to be 3 l/s to determine the number of parallel modules in each set in the first arrangement. Table 1 shows the calculated results. For example in the first stage of the first arrangement we have nine sets of parallel modules and eight parallel modules in each set and the remained four modules constitute the last set. Now the number of booster pumps also could be determined. The first stage of the second arrangement consists of three sets of parallel modules with 48, 20 and 8 modules. Based on the experimental results and assumptions made in this section, required membrane area, number of modules and pumps are represented in Table 1.

Table 1
Specifications of the membrane system.

Stage	Arrangement 1		Arrangement 2	
	First	Second	First	Second
Membrane area (m^2)	2801.38	1114.53	2801.38	1114.53
Number of modules	76	31	76	31
Modules arrangement	9 × 8 + 4	10 × 3 + 1	48 + 20 + 8	20 + 8 + 3
Number of main pumps	1	1	1	1
Number of booster pumps	1	2	N/A	N/A

Energy consumption in RO system is mainly by high-pressure pumps. The typical efficiency of high-pressure piston pumps varies from 50% in small ones to 90% in bigger pumps [53]. We assumed the efficiency of the pumps to be 50%. The power needed for the pump could be calculated from the following equation:

$$W = \frac{Q \cdot \Delta P}{\eta} \quad (3)$$

where W is the power (W), Q is the flow rate (m^3/s), ΔP is the pressure difference (Pa), and η is the efficiency of the pump. The energy consumption of all the pumps could be calculated in this way. We used the average permeate flux of each stage for all its modules to obtain the flow rate of booster pumps.

To compare the membrane process with evaporation, we considered a 4-effect evaporator. In the classical concentration this evaporator should concentrate sugar thin juice from brix 15–60°. But in the hybrid process it should concentrate the product of RO system with brix 20° to the final brix of 60°. In a multi-effect evaporator with N stages, each kg of steam evaporates N kg of water approximately. The evaporated water in the concentration process and the required steam may be calculated with a sugar mass balance. Assuming the latent heat of water evaporation in the input steam conditions (2.5 bar and 137 °C at Bisotoun sugar factory) to be about 2200 kJ/kg, the minimum required energy, i.e. without respect to the real energy consumption of boilers, ancillary equipments, pumps etc., could be calculated based on the following equation:

$$q = m_S\lambda \quad (4)$$

where q is the thermal power (kW), m_s is the mass flow rate of the steam (kg/s) and λ is the latent heat of water evaporation (kJ/kg). The total energy consumption of the 4-effect evaporator and the hybrid RO-Evaporator process with two arrangements is calculated

Table 2
Comparison between energy consumption for reverse osmosis and evaporation.

	Energy consumption (kW)	Energy saving (%)	Substitution coefficient (MJ/kW h)
RO (arrangement 1) + evaporator	184 + 6376 = 6560	33	66
RO (arrangement 2) + evaporator	177 + 6376 = 6553	33	68
Evaporator	9740		

and represented in Table 2. Obviously, a huge amount of energy can be saved by applying membrane system for pre-concentrating the sugar juice. Calculating the required energy shows a 33% reduction in energy consumption of the concentration process.

The benefits may be estimated using the method of the "substitution coefficient" introduced by Electricité de France. This coefficient compares the primary energy saved to the electrical energy consumed in cycles that utilize electricity-consuming operations in substitution of conventional thermal operations [54]. The substitution coefficient is defined by the ratio between the primary energy (thermal) saved in the new process with respect to the conventional process and the amount of electrical energy consumed, relative to the conventional process:

$$CS = \frac{C_1 - C_2}{E_2 - E_1} \tag{5}$$

where CS is the substitution coefficient, C is the consumption of thermal primary energy (MJ or Mcal), E is the consumption of electrical energy (kW h), and 1 and 2 denote the indexes for the conventional and innovating process, respectively.

Taking into account that one kW h of electrical energy requires a power station to burn about 10.5 MJ of primary energy from a combustible source (oil, gas, coal, etc.), the substitution is acceptable when the CS is greater than 10.5 MJ/kW h (2.5 Mcal/kW h). This method is more appropriate for energy saving calculations when a membrane system is compared with a thermal process [54,55]. The substitution coefficient for the RO system with the first arrangement is shown below. We considered the electrical energy consuming in evaporation process as negligible:

$$CS_1 = \frac{35064.15 - 22954.84}{183.94 - 0} = 65.83 \tag{6}$$

This number is greater than 10.5 MJ/kW h. Furthermore, using energy recovery systems like Pelton wheel and Turbo charger to exploit final high-pressure product, may result in higher energy saving [56]. Since the calculations are to show the energy saving potential by membrane processes, the results should be considered as an estimation. Therefore uncertainty values are not meaningful for presentation the results.

The only advantage of evaporation over RO is separation of water without any sugar loss through evaporated water. There are several solutions for this problem in RO system. We may recycle the permeate of the second stage, into the diffusion section of sugar manufacturing plant. Another possible solution is employing a third stage in the membrane system. Due to the experimental results this stage could reject the remained sugar and produce a pure water. Furthermore, the required membrane area will be little due to the low concentration and flow rate of the feed and high permeate flux. The membranes durability will be more and the fouling phenomenon is less and so this stage will not increase the process expenditures too much. The third solution is dividing the permeate stream of the second stage to two or more streams. Because of the wide range of feed concentration in the second stage (brix 4–15°), the permeate of first modules are more pure than the last ones. So we can use this water as the pure water and recycle the permeate of the last modules to diffusion section.

4. Conclusions

In this study, a two-stage RO system with two arrangements for sugar juice pre-concentration has been evaluated for energy consumption. Calculations showed a significant energy saving (33%) by applying this system to increase the brix degree of the thin juice from 15° to 20° prior to final concentration in evaporators. Although the final permeate sugar content is not zero it is negligible. Nevertheless, there are solutions for this problem such as using this permeate in diffusion section extracting sugar from the beet peels or applying a third membrane stage to result in zero sugar in the final permeate.

Acknowledgment

The authors wish to thank the Bisotoun Sugar Factory (Kermanshah, Iran) for providing facilities to run the experiments.

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